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Cincinnati, Ohio

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Advanced Waste Treatment Systems At The EPA-DC Pilot Plant*

By

Dolloff F. Bishop, Thomas P. O'Farrell, John B. Stamberg and J. Winston Porter**

The research at the EPA-DC Pilot Plant in Washington, D. C. is a cooperative effort between the Environmental Protection Agency's Advanced Waste Treatment Research Laboratory and the District of Columbia's Department of Sanitary Engineering. The research goals of the pilot plant are to develop design criteria for advanced municipal treatment plants on the Potomac Estuary and to develop and automate advanced waste treatment technology for nation-wide pollution control and water renovation.

The pilot complex, located at the 250 million gal./day District of Columbia Water Pollution Control Plant, includes a 200,000 gal./day multipurpose pilot plant, a 100,000 gal./day oxygen activated sludge pilot plant, a 50,000 gal./day physical-chemical treatment pilot plant, a 50,000 gal./day ozone pilot plant, a solids handling pilot plant (currently under construction), and supporting laboratory and analytical facilities.

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Pilot Plants

The multipurpose pilot plant (Figure 1) consists of two

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parallel treatment systems with each system containing primary sedimentation, diffused and/or mechanical aeration, and secondary sedimentation. Grit removal, air flotation clarification, and air flotation thickening, are shared between the two parallel treatment systems. Each system is designed for gravity flow at a nominal capacity of 100,000 gal./day and includes an automated flow control system which may be programmed to simulate any diurnal flow variation within the range of 50,000 to 200,000 gal./day.

The primary clarifiers in each system may function as conventional settlers or as chemical flocculator-clarifiers. The processes after primary treatment may receive either pilot plant primary or the main D. C. plant primary effluents.

The two aeration tanks in each parallel system permit operation of essentially all types of activated sludge processes including conventional plug flow, modified aeration, step aeration, contact stabilization, nitrification, and totally mixed aeration. Air is supplied by porous diffusers with up to 3 cu. ft. of air per gal. of wastewater at the maximum capacity of 200,00 gal./day through each of the treatment systems or by mechanical aerators in the totally mixed aeration system. The removable tank baffles and the flow controls provide wide variations in detention time with a maximum of 7.6 hours for series flow through the two aeration tanks at the design flow of 100,000 gal./day. The secondary sedimentation units in each of the parallel systems consist of two peripheral feed clarifiers equipped with rotating hydraulic syphons for rapid sludge withdrawal. The single air flotation clarifier may be employed with either parallel system to

separate the activated sludge from the effluent under aerobic conditions.

Each treatment system in the multi-purpose plant has the flexibility to allow single stage chemical clarification of the raw wastewater. One system has been modified to provide two-stage lime clarification of the raw wastewater. The unused aeration capacity in the modified system has also been converted to pure oxygen activated sludge in one aeration tank and to denitrification in a portion of the other. Thus with the available tankage, conventional aeration, pure oxygen activated sludge, nitrification and two stage chemical clarification of raw wastewater may all be simultaneously operated at 100,000 gal./day. Denitrification, however, may be operated only at 50,000 gal./day.

The 100,000 gal./day oxygen activated sludge pilot plant (1) designed by the Linde Division of Union Carbide (Figure 2) was installed within the multipurpose pilot plant in May 1970. It includes four contacting stages with a total detention time of 1.95 hours at a nominal 100,000 gal./day flow. The four stage reactor permits air tight co-current contacting of the mixed liquor and the oxygen stream. The co-current contacting between stages with oxygen recirculation within each stage produces parallel decreases in oxygen concentration in the overhead gas and in the BOD (5-day biological oxygen demand) of the wastewater. Thus co-current contacting of oxygen and mixed liquor permits efficient utilization of oxygen ($\sim$90%) with high dissolved oxygen even at solids concentrations of 7,000 mg./l. in the mixed liquor.

In the pilot complex, the physical-chemical pilot plant (Figure 3) was designed by EPA for tertiary treatment of secondary effluent and was placed in operation on D. C. secondary effluent in September 1967. The original plant included two-stage lime clarification with intermediate

recarbonation, air stripping for ammonia removal, filtration, pH control, granular carbon adsorption, chlorination, and a sludge processing and chemical recovery system. Breakpoint chlorination has been installed in the pH control basin and a selective ion exchange process employing clinoptilolite for ammonia removal is now being constructed. The ion exchange process may be operated either before or after carbon adsorption.

The processes in the physical-chemical pilot plant have a nominal capacity of 50,000 gal./day with a diurnal flow variation of 4:1. The physical-chemical pilot plant is highly instrumented (2) and features a central control panel from which the plant operator can measure and control all flows, plant pressures, backwash cycle schedules, and many other variables.

The clarification system removes particulate organics and the chemically precipitated phosphorus (3) from the wastewater. This system, designed for the two-stage excess lime process, includes two identical up-flow reactor clarifiers operated in series with an intermediate recarbonation tank. The two clarifiers may also be operated in parallel for treatment of the wastewater by lime-soda ash, alum or iron salts in single stage clarification.

If lime precipitation is employed, the high pH of the effluent from either the first or second lime clarifier permits the removal of ammonia from wastewaters by air stripping (4). The multi-stage air stripping system consists of five cross-flow cooling towers in counter-current operation. Air stripping is not used in cold weather due to low efficiencies or freezing.

The filtration system, with four gravity-flow filters operated

in parallel, has been divided into two operating units. Each two-filter unit may treat either lime clarified secondary or raw wastewater. The "dual media" in the filters consists of 18 in. of 0.9 mm. granular anthracite on top of 6 in. of 0.45 mm. sand. Each filter processes 25,000 gal./day at a surface loading rate of 3 gal./(min.)(sq.ft.) and can receive a maximum of 100,000 gal./day at a surface loading rate of 12 gal./(min.)(sq.ft.). An automatic backwash and surface-wash system permits a maximum backwash rate of 20 gal./(min.)(sq.ft.) which produces 50 percent expansion of the media.

After filtration, the water (clarified secondary or clarified raw wastewater) may be neutralized with carbon dioxide in a stabilization tank. This prevents calcium carbonate scaling and improves the subsequent carbon adsorption process. Alternately, the water may be chlorinated to oxidize ammonia to nitrogen gas before adsorption.

After pH control or chlorination, there is a five column, series operated, granular activated carbon system for removing most of the soluble organics from the water. The columns provide a maximum empty-bed contact time of 35 min. at a surface loading of 7 gal./(min.)(sq.ft.). The carbon system includes automatic backwash and surface-wash systems. A piping manifold permits taking loaded columns off the influent end of the train and bringing regenerated columns on the effluent end in countercurrent operation. Finally, the effluent from the adsorption process may be disinfected by the addition of chlorine.

The sludge processing system in the physical-chemical pilot plant includes a thickening tank, a vacuum filter or a centrifuge and a six-hearth furnace. After dewatering, the sludge is automatically weighed

and fed to the furnace. The furnace, at temperatures up to 1875^{o}F, incinerates the organic solids and calcines the calcium carbonate in the sludge for reuse (5).

The ozone pilot plant was designed, installed and is being operated by the Air Reduction Company through a Federal contract. At the completion of the contract, the ozone pilot plant will have treated various wastewater effluents from the other pilot facilities including secondary, lime clarified and filtered secondary, lime clarified and filtered raw wastewater and carbon column effluents.

The ozone system consists of six reactors (Figure 4) each with ten min. of detention time when operated in series at 50,000 gal./day. Parallel streams of ozone-oxygen gas are injected into the wastewater in a 1 in. pipe at the top of each reactor. Most of the ozone transfer occurs in the reactor inlet pipe and oxidation of the organics continues during the 10 min. detention time.

The oxygen gas residual ozone stream from the top of each reactor is then recycled through a compressor and drier to the ozonator. The recycled oxygen loop provides efficient use of oxygen and prevents ozone leakage to the environment. Ozone generation with oxygen is nearly twice as efficient as with air. The wastewater entering the process is vacuum degassed to prevent build-up of nitrogen in the recycled oxygen stream. Control of the influent water pH is also available ahead of the reactors.

The solids processing system (Figure 5), designed by the Environment Protection Agency and currently under construction by the District of Columbia through a Federal grant, consists of gravity or air

flotation thickening, parallel dewatering processes of vacuum filtration or centrifugation, and incineration or recalcination in a fluid bed furnace. Start-up is scheduled for August of 1971.

The system is sized to handle, in an eight day, the organic or chemical sludge production of any of the 100,000 gal./day pilot treatment systems. The furnace is designed to permit study of incineration of organic sludges, recalcination of lime sludges, and possible regeneration or production of powdered activated carbon. A special solids injector mounted in the top of the fluid bed reactor will properly distribute organic solids, lime or other chemical sludges, or powdered activated carbon into the furnace. The installation of the Solids Process Pilot Plant will complete the pilot systems planned for the EPA-DC Pilot Plant.

These pilot facilities permit evaluation of two basic treatment approaches, physical-chemical treatment and conventional (biochemical) treatment. While the processes in the pilot plant may be arranged to evaluate many variations of the above two approaches, four combinations of processes are currently being studied: These are Physical-Chemical Treatment of Raw Wastewater (PC), Conventional-Tertiary (Physical-Chemical) Treatment (CT), Conventional-Tertiary Treatment with Nitrification-Denitrification (CTND), and Nitrification-Denitrification with Alum Addition (NDA).

The details of the individual processes within each combination are described in subsequent papers. This presentation summarizes the treatment capabilities and the estimated costs for the four combinations. Design criteria for the cost estimates are included.

Physical-Chemical Treatment

In the EPA-DC Pilot Plant, Physical-Chemical Treatment (6) of raw wastewater (Figure 6) consists of two-stage lime precipitation (7) to remove solids and phosphorus, filtration to remove residual turbidity, pH control, ion exchange with chlorination (8) to remove ammonia, and granular carbon adsorption (9) to remove soluble organics. Breakpoint chlorination of the entire ammonia concentration in the raw wastewater may also be applied in the pH control tank.

Ten months of continuous operation on the D.C. raw wastewater from March 1970 to December 1970, usually with a 2.3:1 maximum to minimum diurnal variation but with rain flows with more than 4:1 maximum to minimum variation, produced residuals (Table 1) averaging 6.2 mg./l. of BOD (95% removal), 15.5 mg./l. of COD (chemical oxygen demand) (95% removal) and 0.13 mg./l. of phosphorus (99% removal). Laboratory ozonation (10) of the carbon column effluent further reduced the COD to 5 mg./l. and the BOD to less than 1 mg./l.

While the physical-chemical plant operated continuously without upset, the adsorption columns developed heavy anaerobic growths on the carbon. The growths produced H_2S in the product water and, during backwash, carried excessive amounts of carbon into the backwash decant tank. Because of the biological slimes associated with the carbon, much of the carbon was washed from the decant tank and lost from the system. Carbon recycle from the decant tank to the backwashed column was not employed and carbon losses of from 12 to 25% occurred per replacement cycle.

Improvements in operation and in decant tank design and addition of carbon recycle after backwash are planned to reduce carbon losses. If

high carbon losses continue, the carbon column may be converted to upflow expanded beds to eliminate or reduce the need for backwashing.

The solids processing associated with the physical-chemical treatment sequence cannot be fully studied until the solids handling pilot plant currently under construction is completed. Recent preliminary studies indicate that centrifugation will satisfactorily separate the calcium from the organic-calcium hydroxyapatite components of the sludge. The processing requirements for the centrate with the organic calcium phosphate solids, however, are not known.

An ion exchange process to remove ammonia, provided to the pilot plant by Battelle-Northwest (11), required more operator surveillance than was normally available. Thus, mechanical difficulties prevented continuous operation and reduced process efficiencies. With the exchange unit located ahead of adsorption, five months of intermittent operation reduced the ammonia concentration to approximately 3 mg./l. and the total nitrogen to 5 mg./l. (77% removal). Similar performance occurred with the exchange process relocated after the adsorption system. When the ion exchange system received the necessary operator surveillance, however, the process continuously reduced the ammonia nitrogen to less than 1 mg./l. (8).

Laboratory studies (12) and subsequent pilot studies (8) at the EPA-DC Pilot Plant also revealed that breakpoint chlorination of either the ion exchanged wastewater or lime clarified raw wastewater reduced the NH_3-N to less than 0.1 mg./l. and the total residual nitrogen to approximately 2 mg./l. Chlorination added approximately 200 mg./l.of $NaCl_2$ or $CaCl_2$ to the water. Pilot scale continuous operation of both chlorination and ion exchange are scheduled for determining process reliability.

Conventional-Tertiary Treatment

Conventional-Tertiary Treatment in the pilot plant consisted of primary sedimentation, step aeration (or contact stabilization), and secondary sedimentation to remove BOD and solids, followed by two-stage or single-stage lime precipitation and filtration to remove phosphorus. Physical-chemical ammonia removal (Figure 7) was not operated continuously in the actual pilot sequence as the process was identical to that in physical-chemical treatment of the raw wastewater. Pilot scale carbon adsorption or ozonation for additional soluble organic removal also were not included in the sequence because, with efficient biological treatment, residual COD removal was not needed to meet the discharge requirements for the Potomac Estuary.

The Conventional-Tertiary Treatment, except for ammonia removal, was operated continuously, usually with a 2.3:1 diurnal flow variation, from March 1970 to December 1970, at a nominal capacity of 100,000 gal./day in the conventional processes and 50,000 gal./day in the tertiary (physical-chemical) processes. Lime precipitation was operated as a two-stage process from April through June. The single-stage lime-soda ash process (13) with Na_2CO_3 removing the excess calcium ions was employed from July through December.

In November and December, incomplete removal of the excess calcium ions occurred because interrupted Na_2CO_3 addition in the second flocculation zone of the lime reactor produced $CaCO_3$ scaling of the downstream filters. Additional operation is needed to evaluate low temperature removal of excess calcium ions in the lime-soda ash process.

The conventional-tertiary treatment (Table 2) produced average

residuals of 3.3 mg./l. of BOD (98% removal), 21.7 mg./l. of COD (93% removal), and 0.14 mg./l. of phosphorus (as P)(98% removal). While not required for pollution control in the Potomac Estuary, treatment at a laboratory scale with either carbon adsorption or ozonation (10) further reduced the residual COD to less than 5 mg./l. and the BOD to less than 1 mg./l.

Without a specific physical-chemical ammonia removal process in the treatment series, the conventional tertiary treatment surprisingly reduced the total nitrogen by approximately 50% to an average of 11.1 mg./l. as N for the ten months of operation. In the summer, the aeration process produced nearly complete nitrification (85%). A significant removal of nitrogen by denitrification occurred, probably in the secondary settler, and produced total nitrogen removals that exceed 60%.

Removal of ammonia by selective ion exchange or by breakpoint chlorination may be applied to lime clarified and filtered secondary as effectively as to lime clarified and filtered raw wastewater. However, if continuous high removals of ammonia are to be achieved, biological oxidation of ammonia to nitrate within the aeration process must be prevented. With efficient aeration to insolubilize the BOD, appreciable nitrification could not be prevented in the D.C. wastewater in the summer even with mixed liquor solids of 1,500 mg./l. and with the aeration system operated as contact stabilization. Selective techniques to control nitrification, such as continuous chlorination of the mixed liquor, however, were not employed.

Nitrification-Denitrification

Two treatment combinations of conventional-tertiary treatment with nitrification-denitrification are now being evaluated in the EPA-DC

Pilot Plant. One treatment series (Figure 8) consists of primary sedimentation, air or oxygen activated sludge (14) to remove BOD and solids, nitrification and denitrification with methanol (15) to remove nitrogen, and lime precipitation and filtration to remove phophorus, residual BOD and solids. All bio-chemical processes include an activated sludge reactor and a clarifier. In the nitrification process, an air flotation clarifier has been employed to separate the solids from the nitrified mixed liquor.

In the first activated sludge stage, used for removing BOD, either step aeration (3.8 hours detention at nominal flow of 100,000 gal./day) or the oxygen activated sludge process (1.95 hours detention time at 100,000 gal./day) have been used. The effluents from either step aeration or oxygen activated sludge or an equal combination of the two may be fed to the nitrification-denitrification process.

The three-stage activated sludge sequence (nominal activated sludge-nitrification-denitrification) without lime precipitation or filtration was started-up in November of 1970. The lime precipitation and filtration was added in January 1971. Nitrification has been operated since August of 1970.

Thus, long-term performance has not been determined for the complete treatment sequence. However, the carbon and phosphorus removals previously described for 10 months of continuous pilot operation in conventional-tertiary treatment without nitrification-denitrification (Table 2) are valid for the similar treatment sequence with nitrification-denitrification. In addition, in the summer and fall, nitrification with the air flotation clarifier continuously oxidized the ammonia nitrogen in

the wastewater to less than 1 mg./l. residual NH_3-N (15). In October, upsets in the oxygen activated sludge and step aeration seriously reduced nitrification. Recovery of nitrification after the upset was retarded by the lower winter temperature and a wastewater pH of approximately 7. However, while the operation has been erratic, intermittent performance of nitrification-denitrification, even in January with water temperatures of $55^{o}F$, periodically reduced both ammonia and nitrate concentrations to less than 1 mg./l. as N. Thus, conventional-tertiary treatment with nitrification-denitrification can produce residual total nitrogen of approximately 2 mg./l.at the District of Columbia (Table 3), even in winter. Research at the pilot plant has suggested operating and equipment modifications to minimize biological upsets but additional operation is needed to confirm the reliability for continuously producing residual nitrogen concentrations of 2 mg./l.

In a second alternate with nitrification-denitrification, alum addition into the first activated sludge stage replaces lime precipitation for phosphorus removal. Filtration of the denitrified effluent is retained in the treatment series to remove the suspended solids, particulate BOD, particulate phosphorus and particulate organic nitrogen residuals from the plant effluent.

While alum addition in combination with nitrification-denitrification has not been tested in the EPA-DC Pilot Plant, alum addition (16) into step aeration at a 2:1 weight ratio of Al to P followed by multi-media filtration of the step aeration effluent reduced the residual BOD to approximately 5 mg./l. and the residual total phosphorus to about 0.6 mg./l. as P. The filter runs exceeded 24 hours. Pilot studies are scheduled to

determine the amount of additional phosphorus and BOD removal produced by the nitrification and denitrification processes following alum addition in step aeration.

Costs

Cost estimates for 300 million gal./day plants of each of the four treatment combinations were developed from a cost study performed by the Bechtel Corporation based upon conservative design criteria (Table 4) supplied by the EPA. The treatment plants were designed with maximum loadings determined by a 2.1 to 1 ratio of maximum peak flow to average daily flow. The cost estimates were prepared for Washington, D.C. wastewater with unit costs for chemicals, utilities (Table 5), and labor for the Washington area. These cost estimates were conceptual in nature and did not involve extensive investigation of site, construction, or hydraulic details. They were prepared primarily for purposes of comparison of alternative processes, not for use in budgetary decisions.

The costs are based upon June 1970 price levels with an engineering allowance of 10% and a contingency allowance of 20 on the engineering and construction. The capital charges were computed at an annual rate of 8% (interest and amortization). The costs do not include the cost of land.

In addition the costs presented for carbon treatment and for ion exchange for ammonia removal are based upon concepts which have yet to be demonstrated but which offer substantial capital and operating cost savings. First, the carbon and ion exchange capital costs are based on anticipated use of 100 ft. diameter beds, which would reduce capital costs by about one-third as compared to conventional column diameters. Secondly, operating costs for ion exchange are based upon the use of the waste heat from sludge incineration to supply the heat required for efficient regenerant air stripping. The process also includes unpiloted removal of

the ammonia from the stripping air with an acid adsorption step.

The estimated costs for removal of organics and phosphorus are 22.5 cents per thousand gallons for physical-chemical treatment (Table 6) with a plant detention time of 6.6 hours and 21.1 cents per thousand gallons for conventiona tertiary treatment (Table 7) with a plant retention time of 15.6 hours.

The addition of breakpoint chlorination to remove nitrogen increases the cost of both systems by approximately 6.7 cents per thousand gallons to 29.2 cents per thousand gallons for physical-chemical treatment and 27.8 cents per thousand gallons for conventional-tertiary treatment. Mineralization of the water during breakpoint chlorination of the entire ammonia concentration may be eliminated in both treatment series by substituting selective ion exchange with chlorination to remove the residual nitrogen at a cost of 11.5 cents per thousand. Ion exchange increases the cost for physical-chemical and conventional tertiary treatment costs, respectively, to 34 and 32.6 cents per thousand gallons.

Nitrification-denitrification to remove the nitrogen in conventional-tertiary treatment with two-stage lime precipitation and recarbonation costs 11.2 cents per thousand gallons with a total treatment cost of 33.6 cents per thousand gallons (Table 8) and a plant detention time of 26.3 hours. In this treatment series, single stage lime-soda treatment, if selected to minimize plant size, would increase the lime treatment cost by 0.4 cents per thousand gallons and the filtration cost by 0.3 cents per thousand gallons.

Substitution of alum addition for lime precipitation reduced the costs of conventional-tertiary treatment, with nitrification-denitrification

(Table 9) to 31.4 cents per thousand gallons and the plant retention time to 20.3 hours. Substitution of breakpoint chlorination for nitrification-denitrification in the alum addition system produces the lowest cost plant for organic, phosphorus, and nitrogen removal with a cost of 26.4 cents per thousand gallons at a plant retention time of 10.3 hours (19.4 cents per thousand gallons for organic and phosphorus removal).

Summary

Physical-chemical treatment of raw wastewater consisting of two-stage lime precipitation, filtration, pH control and carbon adsorption in 10 months of operation produced an average residual of 6.2 mg./l. of BOD (95% removal), 15.5 mg./l. of COD (95% removal) and 0.13 mg./l. of total phosphorus (99% removal). The physical-chemical treatment cost was estimated at 22.5 cents per thousand gallons. Conventional-tertiary treatment consisting of primary sedimentation, step aeration, secondary sedimentation, lime precipitation and filtration for ten months of operation produced average residuals of 3.3 mg./l. of BOD (98% removal), 21.7 mg./l. of COD (93% removal), and 0.14 mg./l. of total phosphorus (98% removal). The estimated treatment cost was 21.1 cents per thousand gallons.

Alum addition at a dosage equal to approximately a 2:1 weight ratio of aluminum to phosphorus in step aeration with secondary sedimentation and tertiary filtration reduced residual BOD to approximately 5 mg./l. and residual total phosphorus to 0.6 mg./l. as P at an estimated cost of 19.4 cents per thousand gallons.

As additional treatment, laboratory ozonation of the carbon

effluent from the physical-chemical treatment or laboratory adsorption or ozonation of the conventional-tertiary effluent reduced the COD to less than 5 mg./l. and the BOD to less than 1 mg./l. Thus, removals of carbon and phosphorus exceeding 98% were possible for both physical-chemical and conventional-tertiary treatment.

In nitrogen removal, the selective ion exchange process with intermittent operator surveillance reduced the ammonia content to approximately 3 mg./l. and the total nitrogen to 5 mg./l. as N. However, when continuous operator surveillance was applied to the process, the ion exchange reduced the ammonia nitrogen to less than 1 mg./l. Breakpoint chlorination of the exchange effluent in the laboratory reduced the ammonia nitrogen to less than 0.1 mg./l.

Breakpoint chlorination of secondary effluent in pilot tests reduced the ammonia nitrogen to less than 0.1 mg./l. as N. If combined with solids and organic removal (physical-chemical or conventional tertiary), chlorination at an estimated cost of 6.7 cents per thousand gallons, and ion exchange with chlorination at an estimated cost of 11.5 cents per thousand gallons reduced the total nitrogen to approximately 2 mg./l. as N.

Nitrification-denitrification also reduced ammonia and nitrate residuals to less than 1 mg./l. at an estimated cost of 11.2 cents per thousand gallons. With efficient solids and organic removal, the nitrification-denitrification with effluent breakpoint chlorination should also produce total nitrogen residuals of approximately 2 mg./l. All nitrogen removal processes, however, require demonstration of consistent continuous nitrogen removal.

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* These papers are being submitted simultaneously with the present one. The Editor may wish to incorporate publication references.

Table 1. Physical-chemical treatment of D. C. wastewater

	Pollutants, mg./l.			
Effluent	BOD	COD	P	N
Raw	129	307	8.4	21.8
Clarification	24.3	54.7	0.27	15.2
Filtration	20.4	48.5	0.18	14.9
Ion exchange	13.1*	36.5*	0.20*	5.0**
Adsorption	6.2	15.5	0.13	
Removal %	95	95	99	77*

* Intermittent operation for 5 months

** % removal of nitrogen based upon ion exchange effluent

Table 2. Conventional-tertiary treatment of D. C. wastewater

	Pollutants, mg./l.			
Effluent	BOD	COD	P	N
Raw	129	307	8.4	21.8
Secondary	19.7	59	6.7	14.2*
Clarification	4.2	22.7	0.34	12.1
Filtration	3.3	21.7	0.14	11.1
Removal %	98	93	98	50

* Includes partial Nitrification- Denitrification in summer months

Table 3. Nitrogen removal from D. C. wastewater

Process	Residual NH_3-N mg./l.	Residual N* mg./l.	% Removal
Nitri.-denitr.	1	2	91
Ion exchange	1	2	91
Chlorination	0.1	2	91

* Residuals after filtration for particulate nitrogen removal

Table 4. Design Criteria for treatment systems

Process[5]	Treatment Combination		
	CT[1]	PC[2]	CTND[3] (or NDA)[4]
Grit Chambers			
Detention time (av.), min.	14.7	14.7	14.7
Primary sedimentation			
Detention time (av.), hr.	3.1		3.1
Overflow rate (peak), gal./(day)(sq.ft.)	2990		2990
Activated sludge			
Aeration detention time (av.), hr.	3.8		-
Blower capacity (max 3-day), cu.ft/(lb. BOD)	995		-
Alum dose, mg. Al^{+3}/l			15
Sedimentation-overflow rate (peak), gal./(day)(sq.ft.)	1250		1250
Return sludge as percent of flow	75		75
Lime treatment			
Rapid mix detention time (peak), sec.	30	30	30
Lime dose, mg./l.	350	350	225
Lime recovery	(All of 2nd stage sludge)		50%
Return solids as percent of flow	10	10	10
Flocculation detention time (peak), min.	10	10	10
First stage sedimentation overflow rate (peak), gal./(day)(sq.ft.)	1500	1500	1500
Second stage sedimentation overflow rate (peak), gal./(day)(sq.ft.)	2000	2000	2000
Sedimentation detention time (av.), hr.	2.5	2.5	2.5
Recarbonation or stabilization detention time (peak), min.	10	10	10
Soda Dose, mg./l.			175

Process[5]	Treatment Combination		
	C1[1]	PC[2]	CTND[3] (or NDA)[4]
Ferric chloride dose, (mg. Fe^{+3})/l.	4	4	3
Lime & recarbonation thickeners Feed rate (peak), lb./sq.ft./day	20	20	20
Centrifuge feed rate for 36 in. dia. Bowl and 8% feed solids, dry ton/day	96	96	96
Calciner feed rate, wet ton/(furnace-day)	600	600	600
Nitrification			
Aeration detention time (av.), hr.			4
Aeration blowers, oxygen required (peak), mg./l.			94
Aeration blowers, transfer efficiency, %			8
Sedimentation overflow rate (peak), gal./(day)(sq.ft.)			1250
Sedimentation detention time (av.), hr.			3
Denitrification			
Detention time (av.), hr.			3
Methanol dose, mg./l.			55
Sedimentation overflow rate (peak), gal./(day)(sq.ft.)			1250
Sedimentation detention time (av.), hr.			3
Ion exchange			
Number of stages in service	2	2	
Direction of flow	down	down	
Type of flow	gravity	gravity	
Hydraulic loading (peak), (gal./(min.)(sq.ft.)	10	10	
Pressure Drop across bed 1st stage, lb./sq.in. 2nd stage, lb./sq.in.	15 10	15 10	
Empty bed contact time per stage, min.	3.37	3.37	
Time each stage in service, hr.	24	24	
Time in regeneration cycle, hr.	24	24	
Clinoptilolite capacity, (gm. N)/gal.	10	10	
Clinoptilolite bulk density, dry, lb./cu.ft.	50	50	

Process[5]	Treatment Combination		
	CT[1]	PC[2]	CTND[3] (or NDA)[4]
Number of regeneration stages	2	2	
Time per regeneration stage, hr.	4	4	
Hydraulic loading for regeneration, gal./(min.)(sq.ft.)	6	6	
Type of flow for regeneration	up	up	
Wash volume after regeneration, bed volumes	12	12	
Hydraulic loading for washing, gal./(min.)(sq.ft.)	6	6	
Air-to-liquid ratio for ammonia stripping, cu.ft./gal.	200	200	
Hydraulic loading for ammonia stripper, gal./(min.)(sq.ft.)	2	2	
Ratio of ammonia Absorption spray ($(NH_4)_2 SO_4$ with excess H_2SO_4) to gas, gal./cu.ft.	1/480	1/480	
Activated Carbon			
Number of stages		2	
Direction of flow		down	
Hydraulic loading (av.), gal./(min.)(sq.ft.)		3	
Empty bed contact time per stage, min.		20	
Pressure drop per stage, lb./sq.in.		10	
Carbon loading, (mg. TOC)/(gm. Carbon)		0.15	
Carbon bulk density, lb./cu.ft.		27.5	
Carbon loss per regeneration, %		5	
Backwash rate, gal./(min,)(sq.ft.)		15	
Surface wash rate, gal./(min.)(sq.ft.)		3	
Backwash time, min.		10	
Filters			
Hydraulic loading (av.), gal./(min.)(sq.ft.)		3	
Head loss, ft.		9	
Backwash rate, gal./(min.)(sq.ft.)		20	
Backwash time, min./day		10	
Surface wash rate, gal./(min.)(sq.ft.)		3	

Process[5]	Treatment Combination		
	CT[1]	PC[2]	CTND[3] (or NDA)[4]
Chlorination			
Chlorine dose, effluent (peak), mg./l.	5	5	5
Chlorine dose, influent (av.), mg./l.	15	15	15
Contact time (excess), min.	20	20	20
Sludge disposal			
Vacuum filter feed rate, lb./(hr.)(sq.ft.)	3	3	3
Incinerator feed rate, (wet ton)(furnace-day)	400	400	400
Sewage sludge gravity thickener feed rate (max. day), lb./(day)(sq.ft.)	42	42	42
Flotation thickener feed rate (max. 3-day), lb./(day)(sq.ft.)	30	30	30

1- CT - Conventional-Tertiary Treatment

2- PC - Physical-Chemical Treatmemt of Raw Wastewater

3- CTND - Conventional-Tertiary Treatment with Nitrification-Denitrification

4- NDA - Nitrification-Denitrification with Alum Addition

5- Ratio of Maximum Daily Peak Flow to Average Flow was 2.10

Table 5. Unit costs of chemicals and utilities

Item	
Electrical power	10 mills/k.w.h.
Fuel	$0.95/(million B.t.u.)
Lime	$ 24/ton
Alum	$ 85/ton
Ferric chloride	$120/ton
Sulfuric acid	$ 35/ton
Salt (NaCl)	$ 38/ton
Polymers	$ 1/lb.
Methanol	$0.27/gal.
Activated Carbon	$0.26/lb.
Ion exchange mineral	$0.18/lb.
Chlorine	$ 75/ton

Table 6. Costs for physical-chemical treatment **

Process	Cents per thousand gallons		
	Capital	Operating	Total
Pump and grit	0.6	0.5	1.1
Lime	4.2	3.8	8.0
Filtration*	1.0	0.9	1.9
Solids disposal	1.5	2.4	3.9
Adsorption	4.8	2.8	7.6
Total			22.5
Ion exchange & Cl_2	5.2	6.3	11.5
Chlorination	0.6	6.1	6.7

* 10 mg./l. of Cl_2 added to control biological growth

** 300 Million Gallons Per Day

Table 7. Costs for Conventional-Tertiary Treatment*

Process	Cents per thousand gallons		
	Capital	Operating	Total
Conventional	3.8	3.2	7.0
Solids disposal	2.0	2.5	4.5
Lime clarification	4.2	3.8	8.0
Filtration	1.0	.6	1.6
Total			21.1
Ion exchange & Cl_2	5.2	6.3	11.5
Chlorination	0.6	6.1	6.7

* 300 Million Gallons Per day

Table 8. Costs for nitrification-denitrification and lime precipitation**

Process	Cents per thousand gallons		
	Capital	Operating	Total
Conventional	4.1	2.3	6.4
Nitrification	3.2	2.3	5.5
Denitrification	3.0	2.7	5.7
Lime clarification	3.0	5.4	8.4
Filtration*	1.1	0.8	1.9
Solids handling	2.0	2.6	4.6
Chlorination	0.6	.5	1.1
Total			33.6

* Filter aids added
** 300 Million Gallons Per Day

Table 9. Costs for nitrification-denitrification and alum addition ***

Process	Cents per thousand gallons		
	Capital	Operating	Total
Conventional*	4.2	8.2	12.4
Nitrification	3.2	2.8	6.0
Denitrification	3.0	2.7	5.7
Filtration**	1.1	0.8	1.9
Solids handling	1.7	2.6	4.3
Chlorination	0.6	0.5	1.1
Total			31.4

* Includes alum costs

** Filter aids added

*** 300 Million Gallons Per day

FIGURE 1 -MULTI-PURPOSE PILOT PLANT

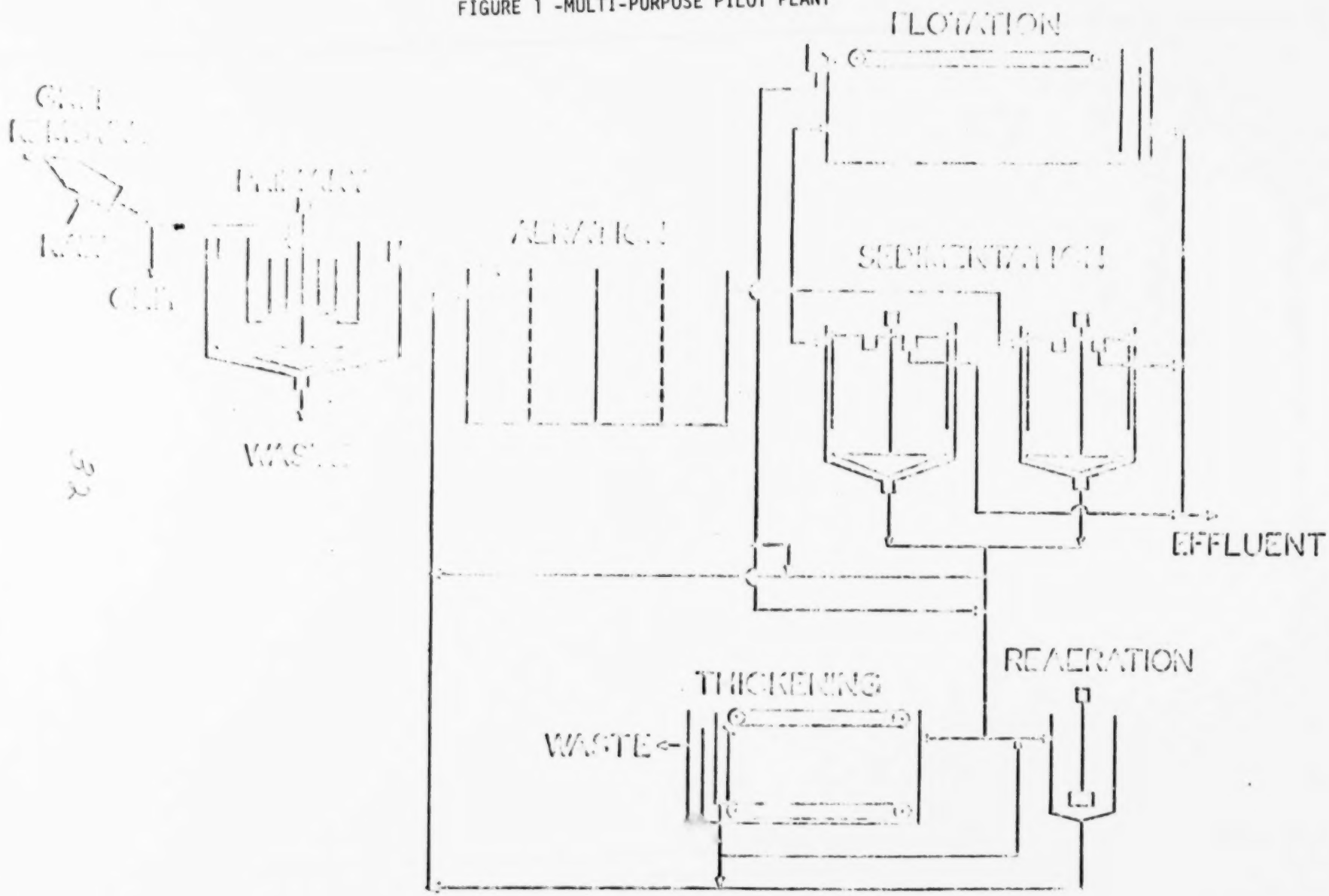

FIGURE 2 - OXYGEN ACTIVATED SLUDGE PILOT PLANT

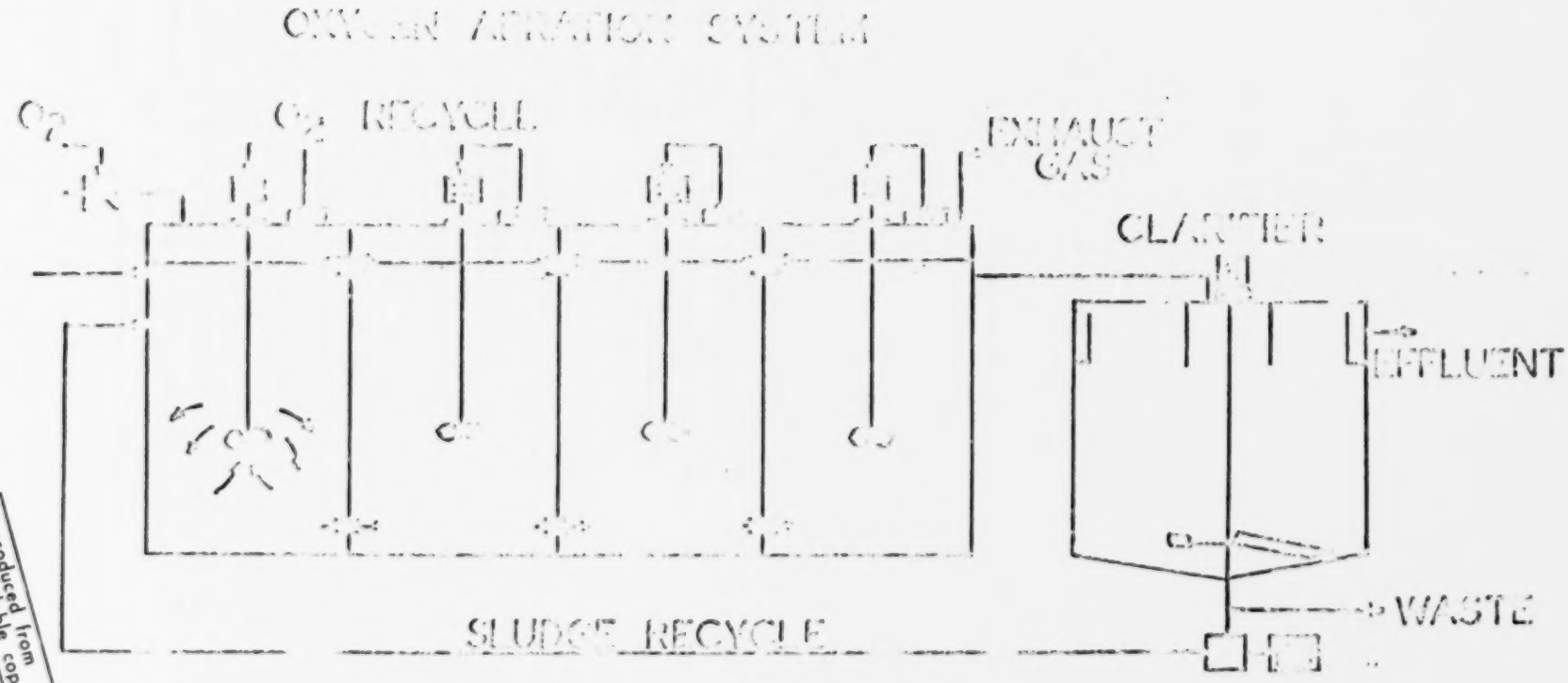

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FIGURE 3 - PHYSICAL-CHEMICAL PILOT PLANT

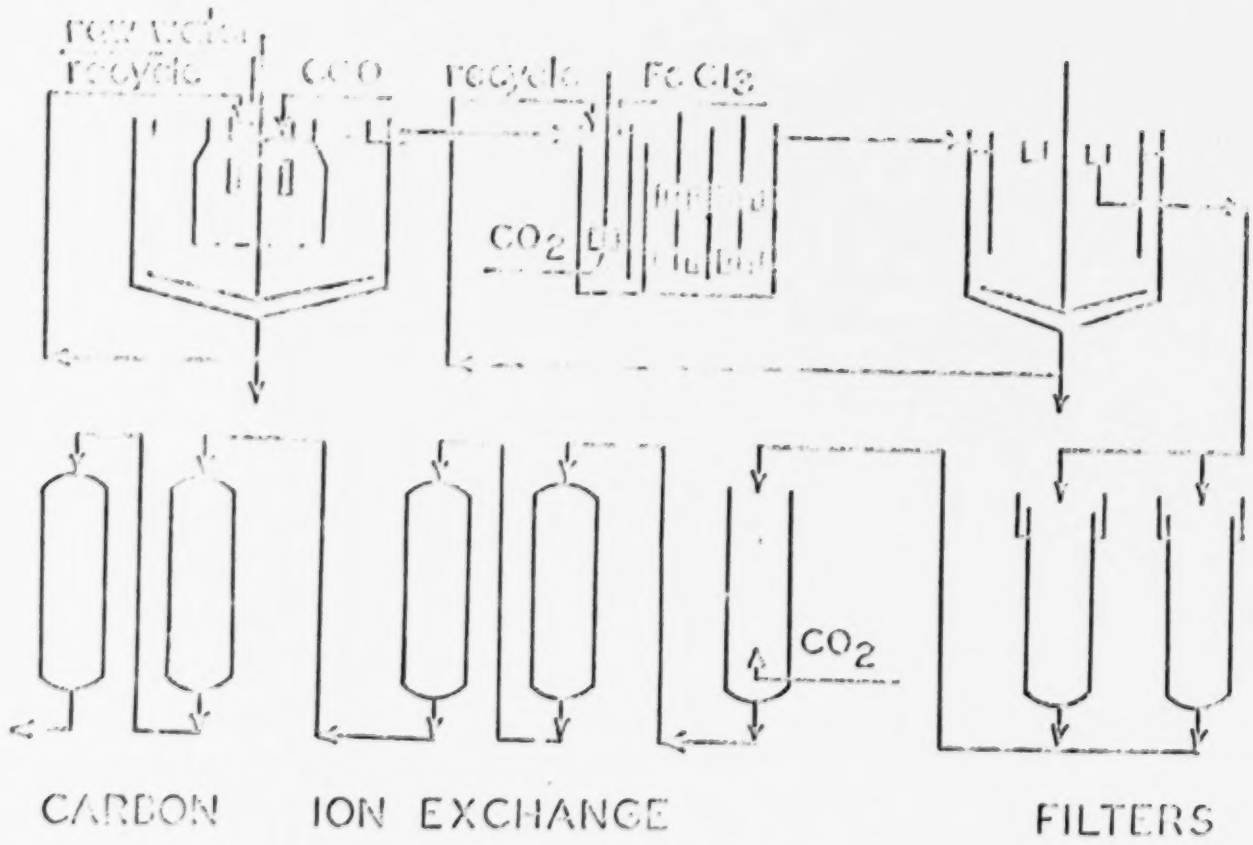

FIGURE 4 - OZONE PILOT PLANT

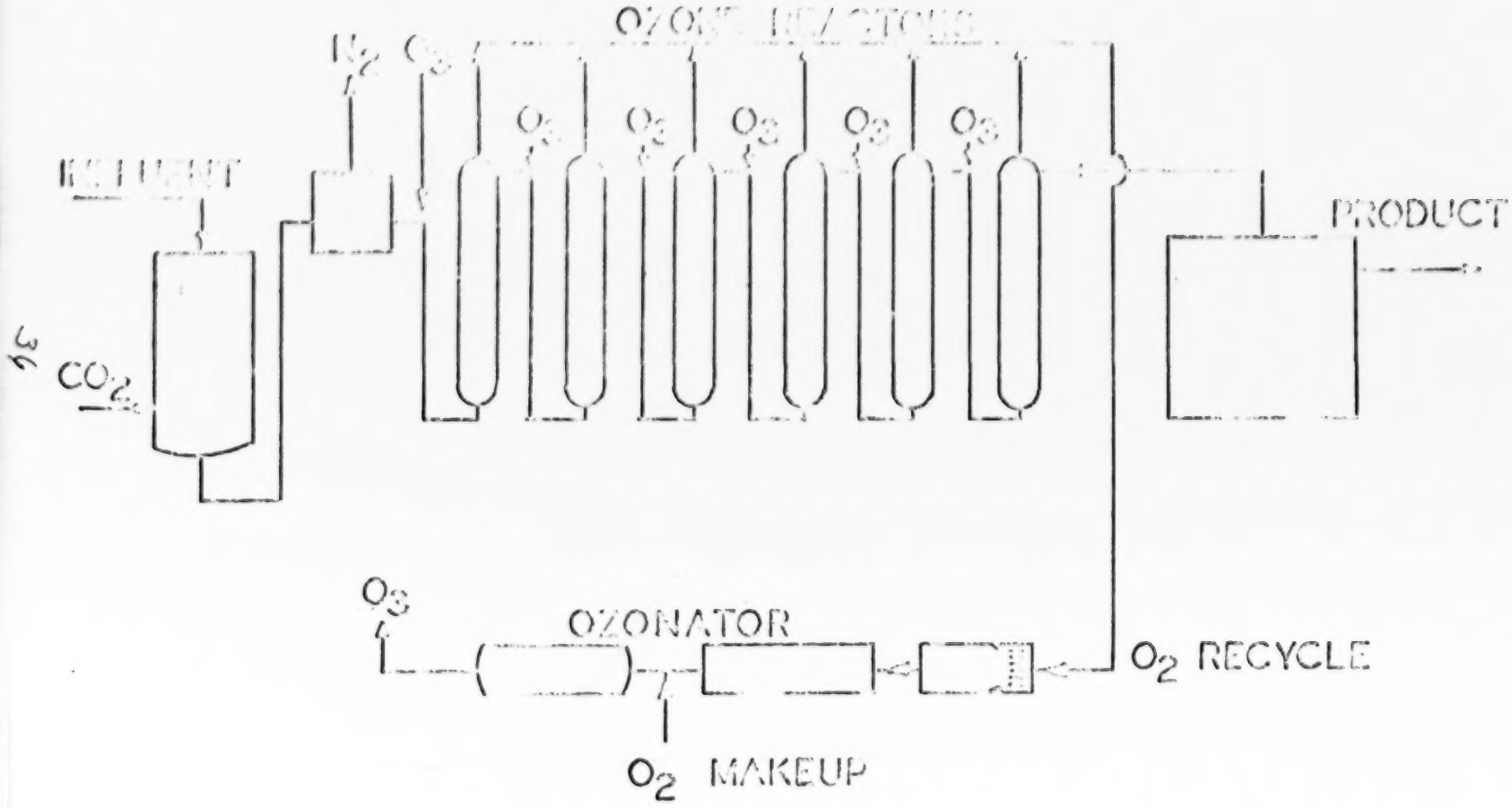

FIGURE 5 - SOLIDS HANDLING PILOT PLANT

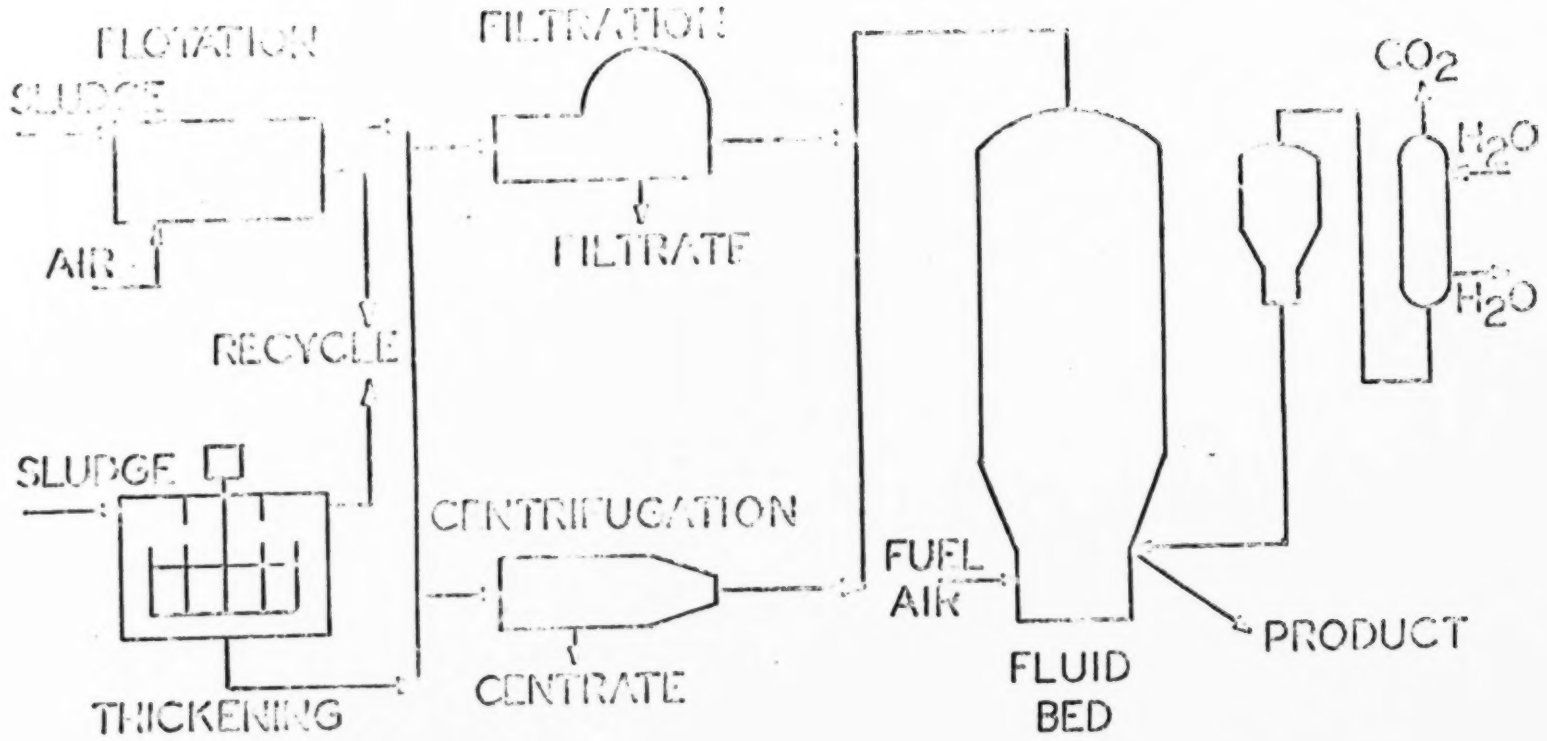

FIGURE 6 - PHYSICAL-CHEMICAL TREATMENT OF RAW WASTEWATER

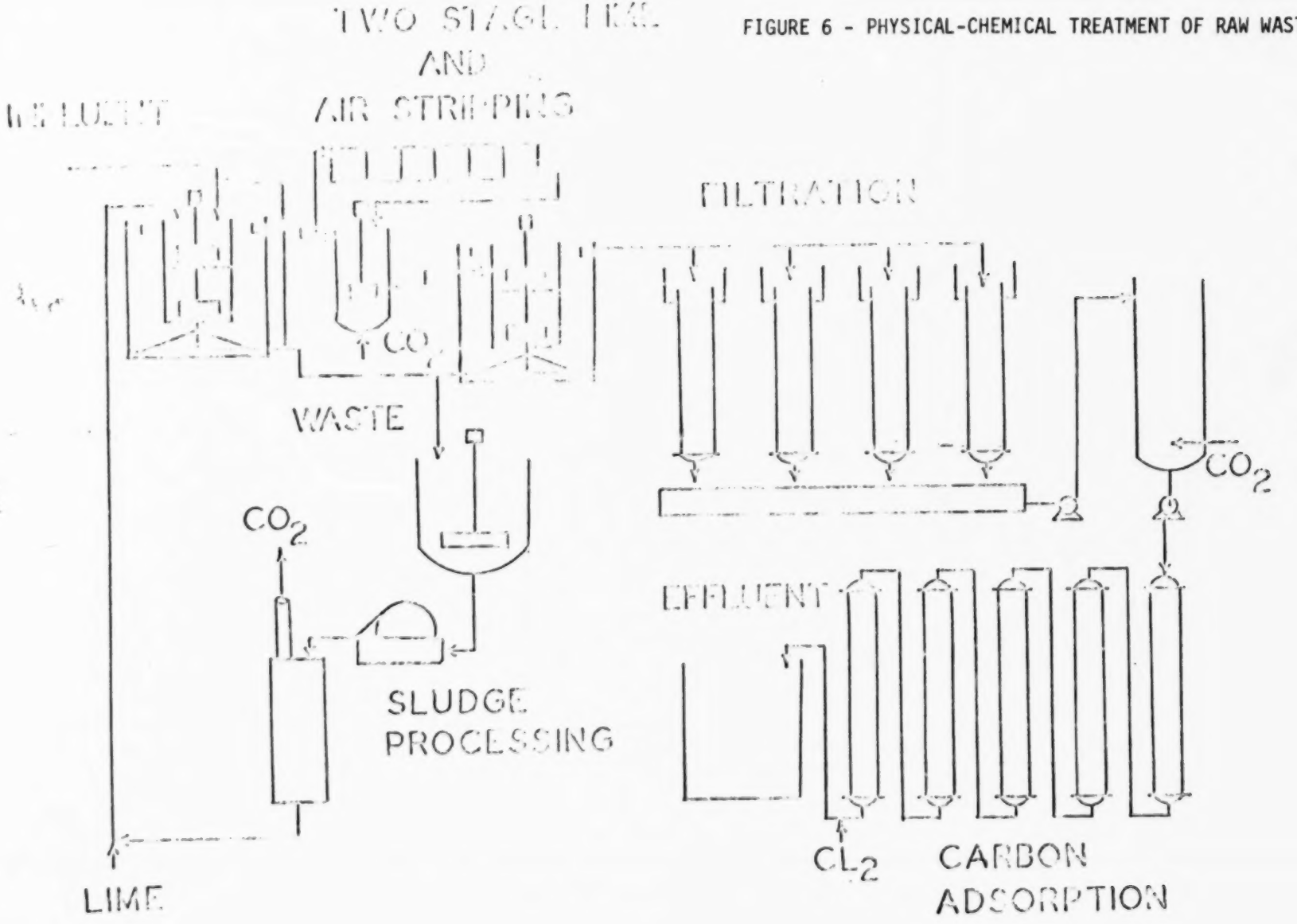

FIGURE 7 - CONVENTIONAL-TERTIARY TREATMENT

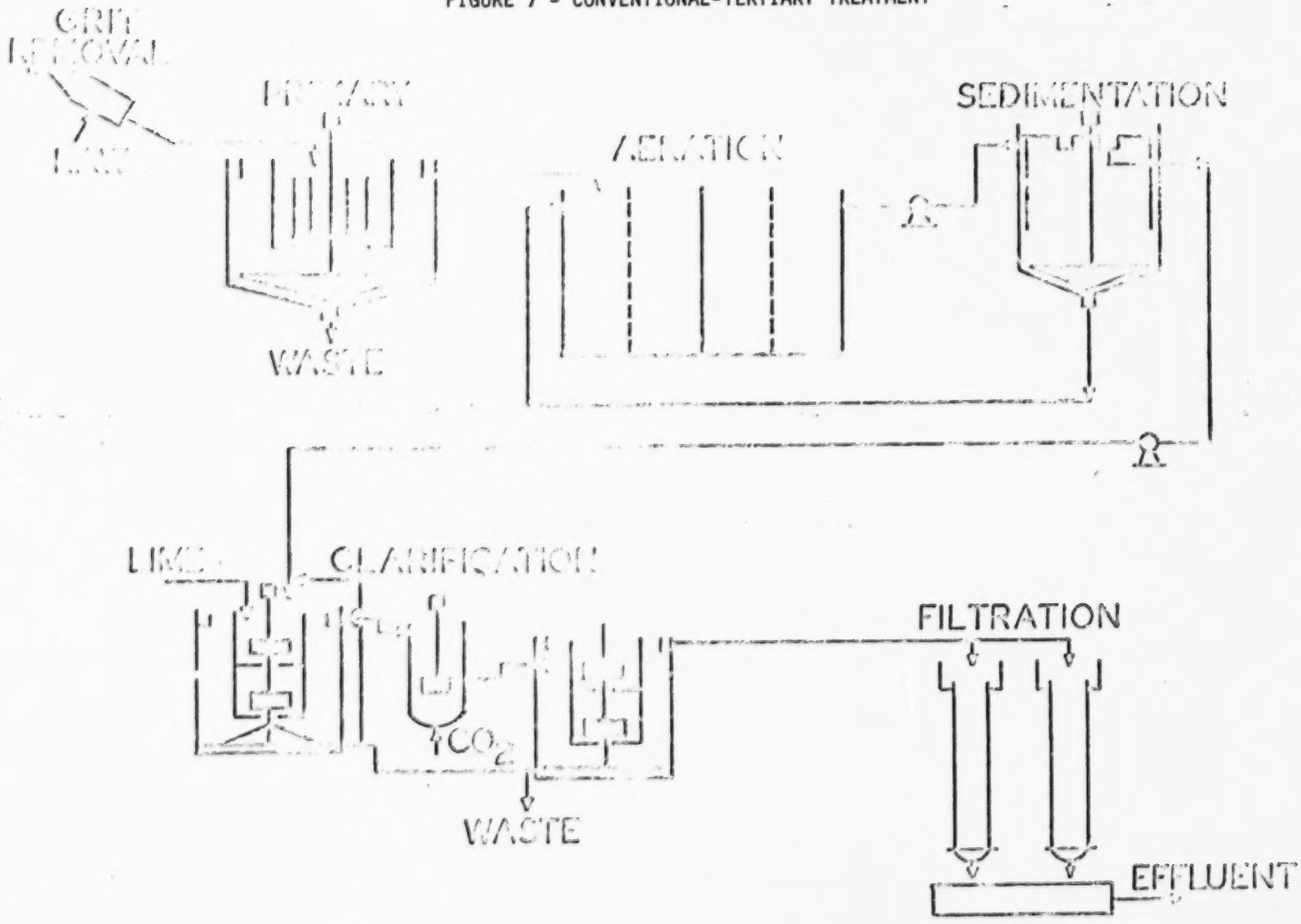

FIGURE 8 - NITRIFICATION - DENITRIFICATION